ECONOMICS OF THE H-COAL® PROCESS

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INTRODUCTION

The escalating cost of energy in the U.S. has stimulated an intensive interest in alternate sources. However, even if major breakthroughs are made in such areas as magneto-hydrodynamics, fusion and solar power, the need for liquid and gaseous fuels for transportation, home heating and existing power plants will be with us until well past the year 2000.

Coal liquefaction offers the potential of substantially reducing the balance of payments deficit while utilizing the enormous U.S. coal reserves which are otherwise environmentally unacceptable. HRI's H-Coal® Process is on the verge of being economically competitive with imported oil, particularly in the central portions of the United States. The studies reported herein start from two basic overall plant integration schemes and then examine the sensitivity of the required fuel oil price to some of the more probable expected variations in process and financial parameters.

H-COAL

The H-Coal Process developed by Hydrocarbon Research, Inc., a subsidiary of Dynalectron Corp., is a direct catalytic hydroliquefaction process. It has been under development since 1963 and has accumulated over 53,000 hours of experimental operation in 25 lb/day bench units and a 3 ton/day Process Development Unit. A 600 ton/day Pilot Plant is currently under construction in Catlettsburg, Kentucky adjacent to the Ashland Oil Co. Refinery. The Pilot Plant project is sponsored by the U.S. Department of Energy, The Electric Power Research Institute, Standard Oil Co. (Indiana), Mobil Oil Corp., Conoco Coal Development Co., Ashland Oil, Inc. and the Commonwealth of Kentucky.

In the H-Coal process, crushed and dried coal is slurried with recycle oils, mixed with hydrogen and liquefied in direct contact with catalyst in an ebullated bed reactor. The reactor effluent is separated into recycle and net product streams in conventional processing equipment. Conversion and yield structure are determined by reactor conditions, catalyst replacement rate and recycle slurry oil composition. The studies reported in this paper are based on an operating severity which produces an all-distillate product. This mode of operations produces a product slate which meets current EPA sulfur specifications without further hydrotreating. Plant size was set at 25,000 TPD coal to the liquefaction section to be consistent with other previously published studies.(1)

In optimizing the overall process flow scheme, the means by which the required hydrogen is manufactured is a very important variable. The two primary alternates are steam reforming of the light gases made in the liquefaction step (a proven process) and partial oxidation of the mixture of ash, unconverted coal and residuum which comes from the bottom of the H-Coal vacuum distllation unit (under development). A second key factor is whether the liquefaction facility purchases power or generates its own. A final significant item is whether there is a customer for the net product gas.

In the present study, two base cases were generated. These are summarized in Table I. Both cases assume on-site power generation. In Case I, the operating severity is adjusted such that the vacuum bottoms, when fed to partial oxidation, put the plant into hydrogen balance. Plant fuel comes from internal streams and net gas is assumed saleable at $2.50/MM Btu. In Case II, the bottoms are carbonized and the resultant coke is fed to the power plant. Excess coke is gasified to prouce a low Btu fuel gas for use in the plant. H2 is produced by steam reforming. As may be seen, the partial oxidation case has a slight economic advantage for the assumptions used. Table II gives the product properties for the two cases. The net gas produced via Case I does not meet interstate pipeline interchangeability specifications. For purposes of this study, the gas was assumed saleable as-is to an industrial customer. If this is not possible, the net gas can be sent to cryogenic purification with C3 and C4 being recovered as saleable liquid products, and a net interchangeable gas being produced with some hydrogen being recycled to the process. The effect of this additional processing can be accounted for in the value assigned to the mixed off-gas as opposed to final product values. This also applies to product gas transporation cost.

SENSITIVITY TO CAPITAL COST ESTIMATE

Because of the many assumptions required for studies of this type, a series of single variable senstitivty analyses were run. The first, and most obviously needed, is the sensitivity to error in the capital investment. Figure I shows the required fuel oil selling price to yield 10% DCF on equity versus percentage change in total capital investment. With gas at $2.50/MM Btu inflation from 1976 to the present appears to give the edge to steam reforming. If net gas can be sold for $3.50/MM Btu, reforming is always the more expensive alternative. This is based on the assumption that bottoms must be utilized on site, by gasification if necessary.

SOURCE AND COST OF POWER

Most of the commercial studies to date have assumed that power must be generated on site. The cases presented herein adhere to this position. There are two main reasons for including power generation in the facility:

1. It is generally assumed that the plant will be located adjacent to a new coal mine. It may, therefore, be impractical, or at least inordinately expensive, to bring in the required power.

2. This facility is estimated to require about 200 megowatts. Even in an industrialized area, this may be more than the local utility can supply.

In order to evaluate the effect of purchased versus generated power, the following assumptions were made:

1. If power can be purchased, gas can be sold.

2. If power can be purchased, carbonized bottoms can be sold. The value of the coke was set using the AGA-DOE guidelines for gasifier chars as 75% of the fuel value of the feed coal: in this case, $0.50/MM Btu.

Figure 2 gives the results of this comparison. The required oil selling price to yield a 10% DCF on equity is plotted against cost of the purchased power at various selling prices for net gas. The horizontal lines represent on-site power generation. As may be seen, reforming with bottoms coke sold at 50¢/MM Btu and partial oxidation with gas worth $2.50/MM Btu both have about the same break even point with purchased power at about 4-1/4¢/Kwh. At $2.50/MM Btu for gas, if power costs less than 4-1/4¢/Kwh, it is always economically attractive to purchase if it is available. If partial oxidation is chosen for H2 generation and the net gas is worth $3.50/MM Btu, purchased power is preferred even if its cost is above 5¢/Kwh.

EFFECT OF PRODUCTS PRICE STRUCTURE

The choice of hydrogen generation processes as well as the decision as to which internal streams should be used as plant fuel are obviously very dependent on the relative value of the various product streams. In Figure 3, the required fuel oil selling price for a 10% DCF return on equity is plotted against naphtha selling price. In addition to the steam reforming case, partial oxidation cases are shown for product gas valued at $2, $2.50 and $3.00/MM Btu, respectively. If the by-product gas is saleable at $2.00/MM Btu or less, steam reforming is the more economical route. With gas valued at $2.50/MM Btu, partial oxidation is preferred to steam reforming when the naphtha value is equal to or greater than the fuel oil.

EFFECT OF COAL PRICE

Coal price is a direct pass through to product price. Because slightly different final product slates (in terms of total barrels per ton) are obtained from the partial oxidation and reforming schemes, coal prices does not affect the two cases in exactly the same manner. Figure 4 shows the required oil selling price versus coal cost for the reforming case and the partial oxidation case with gas valued at both $2.50 and $3.50/MM Btu. With gas at $2.50/MM Btu, reforming becomes preferable at a coal cost at or above $20/ton. With gas valued at $3.50/MM Btu, partial oxidation is preferred.

ECONOMIC MODEL

Because coal liquefaction is very capital intensive, the economic model, in terms of debt/equity ratio, interest rates, DCF and other financial fctors, has a tremendous effect on the required fuel oil selling price. All computations done to this point have used a 55/45 debt/equity ratio, an 8% interest on debt and a 10% DCF return on equity. Figure 5 gives the effect of the debt equity ratio on the required fuel oil selling price. As would be expected, increasing debt ratio decreases the fuel oil price. The quantitative effect is quite pronounced in that a change from 40 to 80% debt decreases the fuel oil price by about \$2.50/Bbl. Figure 6 shows the effect of required DCF[a] return on equity on fuel oil selling price. Again the effect is significant and expectably almost linear. An increase of 2% in the required return on equity at the 45% equity level raises the required oil selling price by about \$1.50/Bbl. These computations reinforce the assertions made by many that the construction of the liquefaction plants is sensitive to the terms and conditions of financing and to taxation policy.

CONCLUSION

These studies show the economic effect of a number of factors which are site specific. Thus, the overall plant confirguration cannot be finally optimized until a reasonably firm location is selected.

COMMERCIALIZATION

A 600 ton per day H-Coal Pilot Plant is currently under construction in Catlettsburg, Kentucky. Operation is scheduled to begin in the first quarter of 1975. The normal commercialization process might wait until Pilot Plant operations were completed before moving ahead. However, the operations on the 3 TPD Process Development Unit have confirmed the operability of the basic process and the real function of the Pilot Plant is equipment testing and fine-tuning of the engineering. Therefore, the commercialization process can be accelerated by immediately beginning such activities as site selection, permit acquisition and preliminary process design. Changes to the yield structure due to the scale difference between the PDU and Pilot Plant will probably not be much greater than the yield variation observed in different batches of coal from the same seam. Therefore, preliminary engineering can begin immediately; this would reduce the commercialization timetable by as much as two years. If such a procedure is followed, a commercial H-Coal plant could be onstream by 1983.

ACKNOWLEDGMENTS

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REFERENCES

1. "H-Coal® Commercial Evaluation", by Fluor Engineers and Constructors, Inc., ERDA Contract No. 49-18-2002.

FIGURE 1. SENSITIVITY TO CHANGES IN CAPITAL

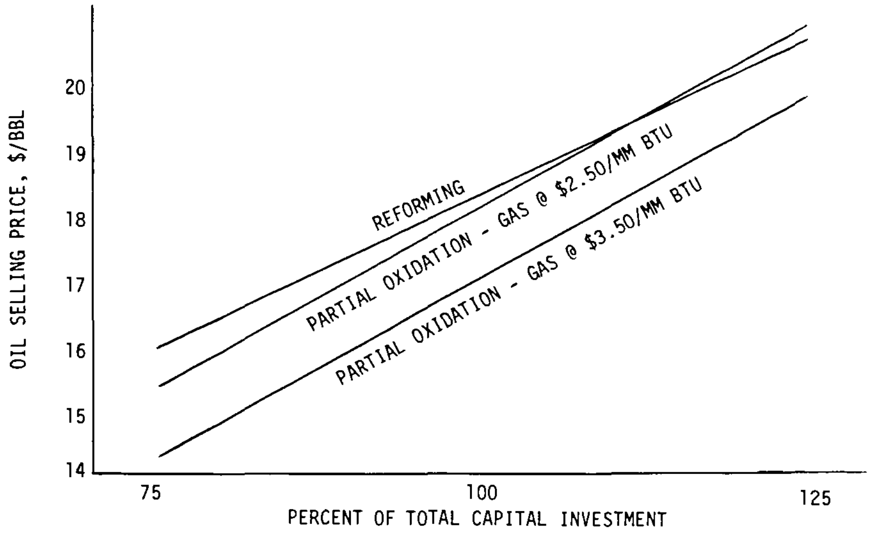

FIGURE 2. EFFECT OF PURCHASED POWER COST

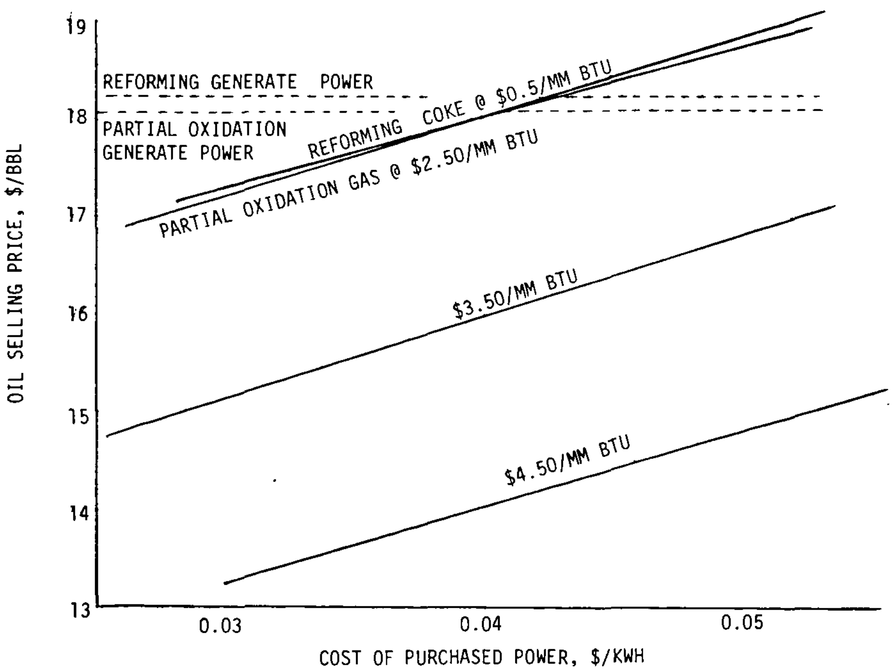

FIGURE 3. EFFECT OF NAPHTHA PRICE AND BY-PRODUCT GAS ON THE ECONOMICS OF DISTILLATE PRODUCTION

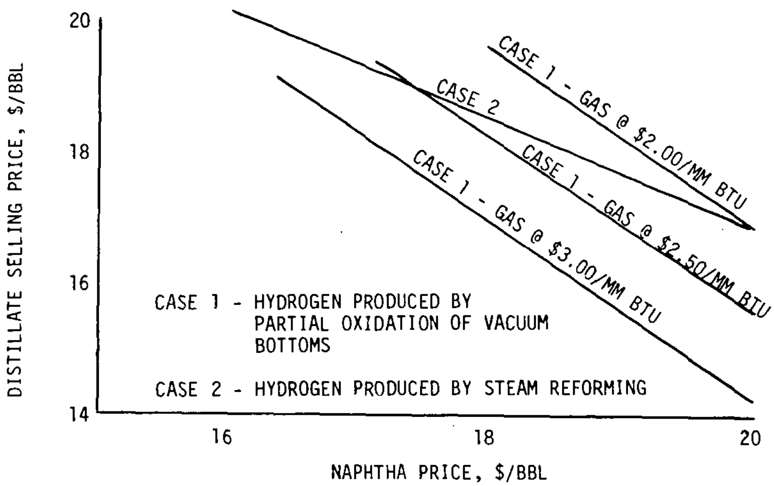

FIGURE 4. EFFECT OF COAL PRICE

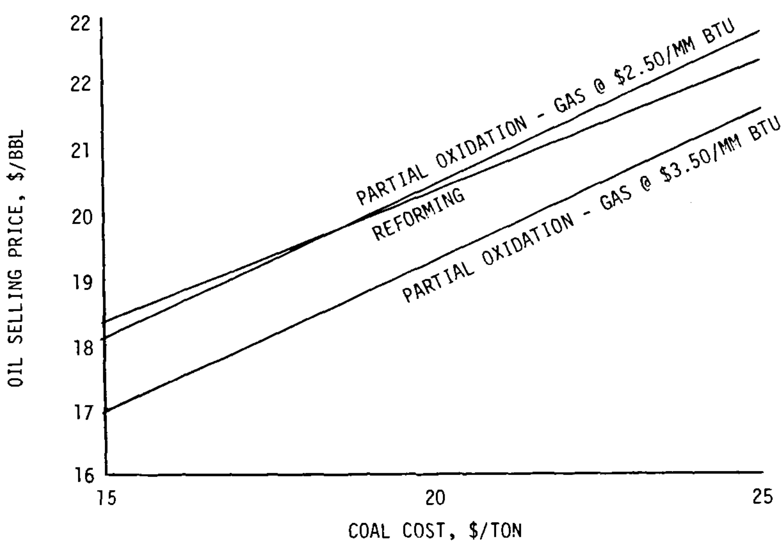

FIGURE 5. EFFECT OF DEBT/EQUITY RATIO

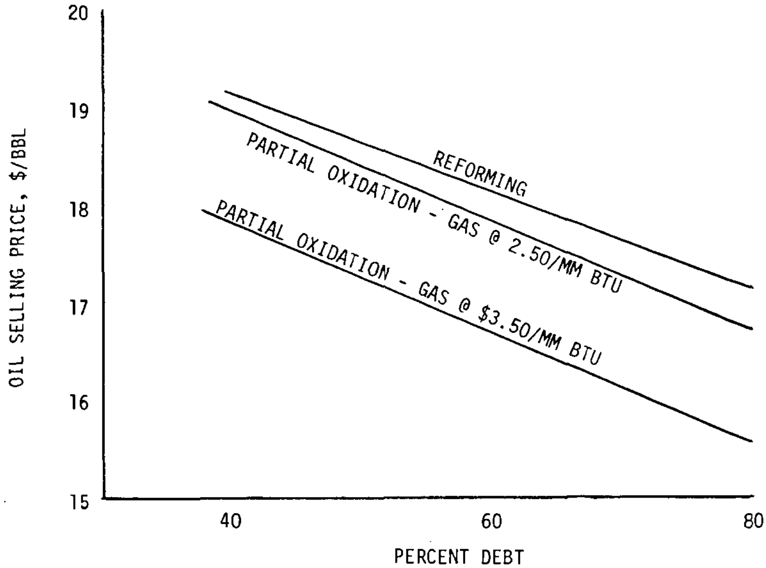

FIGURE 6. EFFECT OF DCF RATE OF RETURN

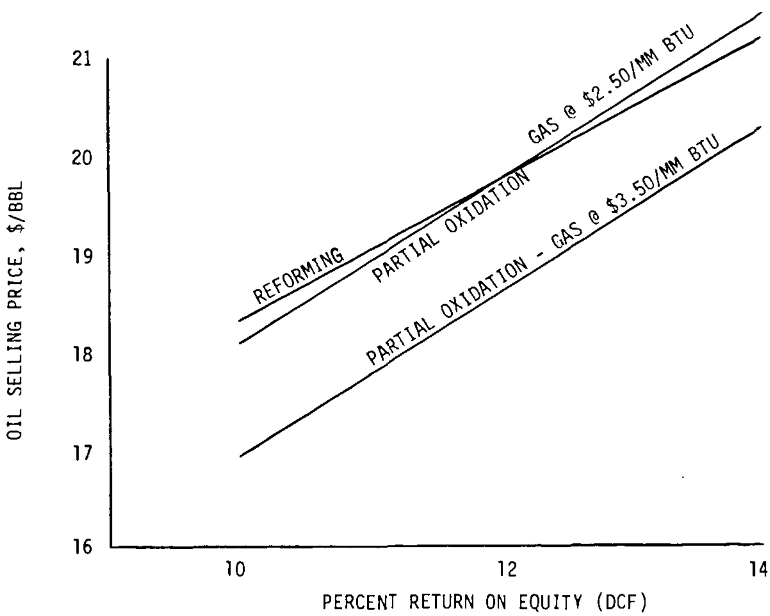

TABLE I

RESULTS TO DATE OF ALL-DISTILLATE

PRODUCTION CASES

Basis: 25,000 T/D coal to hydrogenation
Coal price = $15/Ton, as-received
10% DCF return on equity, 8% interest on debt
Debt/Equity = 55/45
Naphtha value = fuel oil value
By-product gas value = $2.50/MM Btu
Power generation on site
1976 prices used for capital estimates

	Case 1	Case 2
Hydrogen produced by	Partial Oxidation	Steam Reforming
Plant Products		
Naphtha, B/D	35,700	32,200
Distillate fuel oil, B/D	27,200	39,400
Gas, MMM Btu/D (HHV)	70	---
Total depreciable capital investment, $MM	1180	1160
Anhydrous NH_3, ST/D	245	245
Lump sulfur, LT/D	690	708
Thermal efficiency, HHV, %	68.5	67.0
Contribution to Total Oil Selling Price, $/Bbl		
Coal	6.74	5.92
River water	.12	.09
Catalyst and chemicals	.74	.66
Labor, supervision and overhead	.66	.63
Maintenance	1.80	1.56
Insurance and taxes	1.57	1.35
Total Operating Cost	11.63	10.21
Capital-related expense	9.95	8.64
By-product credit	- 3.51	- .59
Total Oil Selling Price	18.07	18.26

TABLE II

SUMMARY OF PRODUCT PROPERTIES IN CASES STUDIED TO DATE

	Case 1 Partial Oxidation	Case 2 Steam Reforming
IBP-400 F Naphtha		
^{o}API	47.0	49.6
Higher Heating Value, MM Btu/Bbl	5.53	5.50
400-975 F Distillate		
^{o}API	21.3	13.5
Wt % Sulfur	0.08	0.12
Higher Heating Value, MM Btu/Bbl	6.13	6.31
Volume %		
400-650 ^{o}F	89.6	63.5
650-975 ^{o}F	10.4	36.5
Gas		
Higher Heating Value, Btu/SCF	1114.0	
Composition, Vol. %		
H_2	28.8	
N_2	2.9	
C_0	2.1	
C_1	37.7	
C_2	16.0	
C_3	7.8	
C_4+	4.7	
	100.0	